HEAT TRANSFER IN A FLUIDIZED BED

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By

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CERTIFICATE

Certified that this work on "Heat Transfer in a Fluidized Bed" has been carried out under my supervision and that this work has not been submitted elsewhere for a degree.

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A C K N O W L E D G E M E N T

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A B S T R A C T

Heat transfer studies in a fluidized bed with and without a chemical reaction (exothermic) were carried out. The chemical reaction carried was between hydrochloric acid gas and ammonia gas. It was found that the heat transfer coefficient increased with the formation of additional solid particles - ammonium chloride. At a constant mass velocity, the bed temperature varied with the amount of heat liberated. The ratio of exit temperature of air to that of the bed was found to be a strong function of the mass velocity.

C O N T E N T S

LIST OF FIGURES

LIST OF TABLES

CHAPTER 1

I N T R O D U C T I O N

The term fluidization signifies the motion of granular particles in the fluid moving upwards. It is due to the motion of particles that increased heat transfer, increased mass transfer, absence of axial and radial temperature gradients occur.

A fluidized bed has the following characteristics :

1. Local temperature and solids distribution are much more uniform than in the fixed bed;

2. Heat transfer coefficients in fluidized beds are higher than in fixed beds operating under comparable flow conditions;

3. Solid particles may undergo attrition or size reduction and there may be equipment erosion.

Heat Transfer Considerations in a Fluidized Bed :

Heat transfer in a fluidized bed may be considered from two different view-points : (a) heat transfer between the particles and the fluid and (b) heat transfer between the fludized bed and a boundary heat transfer surface. Although, the order of magnitude of heat transfer coefficient(*) between the solid particles and fluid in a fluidized bed is almost same as compared to that in a packed bed, yet the particle size in a fluidized bed, narmally encountered is very small (5-100 μ) as compared to that in a packed bed (3mm. & above), the quantity of heats transferred per unit time increases tremendously.

If we compare the order (s) of magnitude of heat transfer co-efficients between the bed and wall for (i) a fluidized bed, (ii) a packed bed, and (iii) air-wall for an empty tube, the current interest in the

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(*) About 8 BTUs/Hr. Sq.Ft.°F. at a mass velocity of 500 lbs/Sq.ft. hr. and about 40, at 1100 lbs/Sq.ft. hr. (12).

fluidized bed heat transfer will be readily understood.

Heat Transfer Equipment	Magnitude
Empty tube	.5 - 1 BTU/hr. sq.ft °F.
Packed bed	3 - 5 BTUs/hr. sq.ft.°F.
Fluidized bed	10 - 100 BTUs/hr. sq.ft.°F.

VAPOR PHASE PRODUCTION OF AMMONIUM CHLORIDE-CHEMICAL REACTION- IN A FLUIDIZED BED :

It was intended to study heat transfer characteristics in a fluidized bed without and with a chemical reaction. Study of a simple exothermic chemical reaction such as above, would give the modifications introduced in the characteristics, if any.

To carry out the ammonium chloride formation from hydrochloric acid gas and a ammonia gas was chosen for the following reasons :

1. the reaction is instentaneous and no kinetics is involved;
2. the reaction is strongly exothermic - the heat of the reaction is + 41.9 K.cal/gm. mole;
3. it can be visually observed whrther the reaction is taking place or not;
4. the product, being in finely solid state, is removed from the reactor by the fluidizing medium i.e. air;
5. reactants are easily available.

In the following pages, work done in this field (without chemical reaction) by many workers and the work done in the present investigation have been discussed.

CHAPTER 2

PREVIOUS WORK IN GAS-SOLID DENSE PHASE FLUIDIZED BED HEAT TRANSFER

MECHANISMS PROPOSED BY DIFFERENT WORKERS :

A lot of work has been done in the field of heat transfer in a fluidized bed in the last three decades. Various authors have proposed different empirical correlations to evaluate the magnitude of heat transfer coefficient - bed to wall. Although, most of them have proposed one mechanism or the other, qualitatively, it is only in a few cases that the empirical correlation has been proposed to substantiate the proposed mechanism quantitatively [3, 7, 12, 15]. In the following lines, some of them have been briefly discussed.

Dow and Jakob's Model (1951) :

The authors [5] have proposed a model. According to this, there exists a narrow thermal layer next to the wall and it constitutes the main resistance to the heat flow. A limited thermal mixing region is there at the bottom of the fluidized bed. The core of the bed in the interior is largely at thermal equilibrium with radial temperature gradient virtually non-existent.

Wicke and Fetting Model (1954) :

According to the authors [12], in contact with the wall, there is a fluid film of thickness δ_G, $0.1\ D_p < \delta_G < 0.5\ D_p$. This is followed by a solids boundary layer of thickness δ_p, 1 to 3 mm. thick. Beyond this layer there is the fluidized core. In the boundary layer, the solids move essentially parallel to the wall; however, there is also lateral solids flow between the boundary layer and the core. Near the fluid film, the particles move dowwards mainly, where-as in the core, they move upwards mainly.

Both of the above models are based on the observation that there is practically no radial or axial temperature gradient in the core.

Van Heerden and Co-workers Model (1953) :

Although, it was established that increase in heat transfer in a fluidized bed is because of the motion of the particles only, previous workers proposed that solid particles act as turbulence promotors without contributing directly to the transport of heat by their convective motion.

The authors[9] proposed that the heat capacity of solid particles per unit volume of the bed is about 1,000 times greater than that of intersticial gas while mean particle velocity is at most ten times lower than the gas velocity. Thus, it is clear that convective transport of heat by the moving particles largely outweighs all other modes of heat transfer. If the heat capacity of the particles is reduced infinitely without changing the other physical properties, the rate of heat transfer will drop to the same value as obtained in the fixed bed of the same particles and the same gas velocity.

The authors proposed that the fludized bed is equivalent to a well stirred liquid in which the intersticial gas only serves as a stirring agent and as a heat transferring medium between the adjacent particles and between particles and the wall. The rapid heat exchange among different parts of the bed is brought about by the turbulent motion of the particles which is completely equivalent to the high eddy diffusivity of a well stirred liquid, and which can be expressed as eddy thermal conductivity (0.6 to .3 BTUs/hr. ft. °F).

Mickley and Fairbank [15] Model (1955) :

On the basis of their experimental finding that the heat transfer coefficients obtained with fluidized beds are proportional to the square root of the quiescent bed thermal conductivity, they concluded that the agencies which control heat transfer may be looked upon as unsteady state of diffusion of heat into the mobile portions of the bed termed "packets". These packets themselves are believed to have a transient existance. Thus, they would tend to disperse and re-form elsewhere. The packets are pictured as contacting the heater wall and thereby facilitating heat transfer. A correlation of heat transfer coefficient along this line will demand the assumption of a so called stirring factor, necessary for moving the particles, joined together as packets, through the bed. In other words, stirring factor signifies the frequency of solid particles, contacting the heat transfer surface. In regard to the stirring factor, the theory demands that it is related to heat transfer coefficient, the thermal conductivity of the quiescent bed, the density and the heat capacity of the packet. The stirring factor itself depends upon the mechanical properties of the bed and particles.

Botteril Model (1961, 1963) :

Concluding from the experiments carried out under almost vacuum the author [2] proposed that in order to carry out heat transfer, the presence of some heat transfer medium i.e. the fluidizing gas, is essential. He further proposed [3] that the conductive heat transfer between a single particle immersed in a fluid and a heat transfer surface can be regarded as the basic stage in heat transfer to gas

fluidized beds of spherical particles where radiative transfer is negligible. The transfer rate is controlled by particle residence at the transfer surface.

Summarising the above discussion, it may be said that the function of the fluid is to stir the particles, act as heat transfer medium between the solid surfaces where there is no mechanical contact, whereas the solid particles act as turbulence propotors as well as heat carriers exchanging the heat amongst themselves, with the gas and with the wall. They also reduce the film thickness. The result is that almost uniform temperature throughout the bed exists and increased heat transfer takes place.

VARIOUS PARAMETERS AND THEIR SIGNIFICANCE :

Inspite of the lot of work done in this field, there does not exist one single theory or a general correlation applicable in each and every case. Moreover, because of extremely complicated hydrodynamic behaviour of a fluidized bed, regrous mathematical treatment has not been possible. That is why, the approach has been empirical in this field.

To understand the effect of different parameters and to propose a mechanism, various authors varied different parameters to see the effect upon resulting heat transfer. The important parameters are :

(i) mass velocity of the fluidizing gas;

(ii) particle diameter,

(iii) heat capacity of the solid particles, and

(iv) thermal conductivity and the kinematic viscosity of the fludizing gas.

The less important ones are : heat capacity of the fluidizing gas, height of the fluidized bed, and tube diameter, etc.

Recently, there has been a trend (6, 17) to introduce a few modifications in the design of the fluidized bed eg. introducing baffles or some other packing material. The aim is to obtain smooth fluidization even at higher values of mass velocity i.e. higher values of reduced mass velocities which otherwise will give rise to slugging, to get higher magnitudes of heat transfer coefficient than would be obtained under normal fluidizing conditions.

Mass Velocity Dependence :

This is the most important factor upon which heat transfer depends. Different workers have given the following dependence of heat transfer coefficient upon mass velocity.

Dow & Jakob (5), for a mechanically smooth fluidized bed	$h \propto G_f^{0.8}$
Van Heerden and Co-workers (8)	$h \propto G_f^{0.45}$
Toomey & Johnstons (18)	$h \propto \log \frac{U_f}{U_{mf}}$
Levenspiel & Walton (14)	$h \propto G_f^{0.3}$
Leva (12)	$h \propto G_f^{0.36}$

In, almost, all the above cases, when h and G_f were plotted an a log-log scale, a curve with a large curvature was obtained and the authors approximated the central portion of the curve with a straight line. It is the general observation that at lower values of reduced mass velocity, G_f/G_{mf}, the dependence of heat transfer coefficient on the former is higher. At the higher values of G_f/G_{mf} the curve becomes more and more flat.

Physically interpretting, it may be said that since the particle motion is an increasing function of the mass velocity, that is why heat transfer coefficient increases with the same.

For generallised correlations and the limits of variables see pp 45-47 (appendix)

Effect of Particle Diameter :

Dow & Jakob [5]	$h \propto D_p^{-0.17}$
Van Heerden & Co-workers [8]	$h \propto D_p^{-0.55}$
Levenspiel & Welton [14]	$h \propto D_p^{-0.7}$
Leva [12]	$h \propto D_p^{-0.04} \; \eta^{0.36}$
Bhat & Weingaestner [1]	no appreciable effect

Since minimum fluidization mass velocity is itself a very strong function of particle diameter and the quality of fluidization, defined in terms of fluidization efficiency, η , is a strong function of reciprocal of particle diameter, it may be said that heat transfer coefficient increases with decreasing particle size. Physically interpretting, it may be said that at any value of mass velocity, the reduced mass velocity being greater than unity, the particles with lesser diameter agitate more vigorously; that is why higher magnitude of heat transfer coefficient is obtained.

Heat Capacity of Solid Particles :

Dow & Jakob [5]	$h \propto (\rho_s \; C_s)^{0.25}$
Van Heerden & Co-workers [8]	$h \propto (\rho_s \; C_s)^{0.36}$
Mickley & Fairbank [15]	$h \propto (\rho_m C)^{0.5}$
Leva [12]	$h \propto (C_s \rho_s)^{0.4}$

$\rho_m C$ is the heat capacity of quiescent bed.

As previously mentioned (Van Heerden Model), heat transfer is mainly because of the convective transport by the moving particles, hence the above trend.

Thermal Conductivity of the Gas :

Dow & Jakob (5)	$h \propto k$
Van Heerden & Co-workers (8)	$h \propto k^{0.5}$
Toomey & Johnstone (18)	$h \propto k$
Leva (12)	$h \propto k^{0.6}$
Mickley and Fairbank (15)	$h \propto k_m^{0.5}$

k_m = thermal conducitivity of the quiescent bed.

Assumption of a laminar film along the wall justify the power of k to be unity. For turbulent conditions and for a packed bed, two-thirds power of k is already well established. So for a fluidized bed, the power with magnitude of 0.5 to 0.6 is fairly justified.

Dependence on Kinematic Viscosity of the Gas :

This is purely because of hydrodynamic conditions. It will effect only the film thickness(12). Practically, in all the correlations, where Reynold's number appears, it is taken care of along with the mass velocity.

Dependence upon the Height of the Bed :

Although many workers have not mentioned the effect of bed height on the magnitude of heat transfer coefficient, it seems that the later will decrease with the increase of former. The reason is that beyond a certain height in a particular case, bed homogeneity and quality of fluidization deteriorate with increasing bed height. So, the heat transfer coefficient will decrease with bed height. However, Dow & Jakob (5) have given the following dependence,

$$h \propto L_f^{-0.65}$$

Heat Capacity of the Gas per Unit Volume :

As already stated, heat transfer takes place mainly because

of the convective transport of solid particles, so, that heat capacity of the gas may not have significant effect on heat transfer coefficient. However, following dependence has been reported :

Dow & Jakob [5]	$h \propto (C_F \; \rho_F)^{-0.25}$
Levenspiel & Walton [14]	$h \propto C_F$
Van Heerden & co-workers [8]	$h \propto C_F^{-0.14} \; \rho_F^{-0.18}$
Mickley & Fairbank [15]	$h \propto C^{0.5}$

Thermal Conductivity of Solid Particles :

Since the solid particles encountered in the fluidized bed are very small, surface area per unit volume is very high. So, it is almost instantaneous when the average temperature of a particle is practically the same as the external surface temperature of the particle. Thus, thermal conductivity of solid particles does not play a significant role.

Effect of Tube Diameter :

Wall effects may be common in fixed beds where large particles relative to the tube diameter, are sometimes used. Because of the larger voidage in the peripheral ring, a somewhat larger proportion of fluid passes through the ring than the core of the tube. Wall effect is commonly expressed as a ratio of the tube diameter to the particle diameter ratio, D_t/D_p. As the ratio increases towards high values, the ratio of ring to total area decreases rapidly. For fluidized systems of the tube normally encountered, the ratio varies from 300 to 1000. For these values, area ratio is virtually negligible.

In short, any parameter will affect the magnitude of heat transfer coefficient which affects particle motion, or the thermal

conductivity of the medium.

[illegible]'S CORRELATION :

The author [12] has chosen Wicke and Fetting model, described previously for developing a correlation. In the development, there are two important considerations. Firstly, the particle velocity and bulk density (or particle population) along the wall influence the thickness of the film. In other words, heat transfer coefficient, being dependent upon the film thickness, depends upon the particle velocity and particle population along the wall.

Secondly, in the buffer zone and central core- which constitute most of the bed - convective heat flow by solid particles takes place; therefore, solid heat capacity has also to be accounted for.

Making certain assumptions regarding the hydrodynamic behaviour of the bed, and making use of the above considerations, the auther proposes.

$$h \propto C_1 \ (f) \ \frac{G_f \eta}{\mu R} \qquad \ldots (1)$$

where C_1 = a constant (to account for the thermal and mechanical properties of the solid particles).

η = fluidization efficiency

R = bed expansion ratio

Whereas, C_1, a constant, takes into account the thermal and mechanical properties of the solid particles, remaining part of the R.H.S. of the above expression has been formulated out of the first consideration, mentioned above.

The data of different workers was analysed according to the above relation. The following correlation was developed after

plotting and cross-plotting ,

$$\frac{h\,D_p}{k} = 0.16\left[\frac{C_s\,\rho_s\,D_p^{1.5}\,g_c^{0.5}}{k}\right]^{0.4} \times \left[\frac{G_f D_p \eta}{\mu\,R}\right]^{0.36} \quad \ldots (2)$$

The equation consists of three dimensionless groups. It essentially relates a Nusselt number with the last group, a modified Reynold's number. The central group accounts for the effect of particle properties on heat transfer coefficient.

MODIFICATIONS IN THE CONVENTIONAL FLUIDIZED BED HEAT TRANSFER SET-UP :

Some workers (16, 12, 17) modified their set ups and then studied heat transfer characteristics. For example, Massimillah & co-workers (12), introduced the baffles. It was found that the maximum observed between h and G_f with a conventional set up completely disappeared and heat transfer coefficient increased continuously with the mass velocity.

J.P. Sutherland and coworkers (17) studied the heat transfer characteristics with different packing materials for example, smooth spheres, rough spheres, Berl-Saddles, Rasching-rings and cylindrical screen packing. It was found that the rate of wall to bed heat-transfer was in most cases adversely affected by the addition of packing materials, because of restrictions of particle movement within the bed and the vessel walls. This was true for spheres, Rasching-rings and Berl-Saddles. However, cylindrical screen packing has been found to cause little or no reduction in maximum heat transfer rates because of the open structure. Its relatively small interference with particle movement combined with its higher porosity and ability to eliminate slugging suggest that this type of packing should have a useful application in fluidized bed, of free flowing solids. With

spherical packing, values of heat transfer coefficient of the order of magnitude of 70% of that in a conventional bed have been obtained.

Some other workers [6] studied fluidization characteristics with the rectangular and square sectioned columns with fins (circular rods) protruding inside the column. They found that slugging was practically completely removed and the whole bed was homogeneous. The authors claim that rate of heat transfer from wall to bed in such system can be increased in accordance with the finned surface and overall heat transfer coefficient would in such cases be far higher than available in conventional fluidized-bed equipment.

EXPERIMENTAL CONDITIONS

Heat transfer characteristics with and without a chemical reaction in a fluidized bed were to be studied. A number of runs were taken and in both the cases; sand particles were fluidized by air. Heat transfer took place from bed to wall.

HEAT TRANSFER WITHOUT A CHEMICAL REACTION :

Experimental Set-up :

The flowsheet is given in Fig.1. The constant and steady supply of air was available from an air-compressor. There were a pressure regulator and a rotameter to regulate the pressure and to meter the flow rate respectively. A needle valve was used before the rotameter to control the flow rate and a mercury manometer, to show the pressure of inlet air. The air was preheated in a preheater and then led to fluidization column.

The preheater consisted of a M.S. pipe with 1.5 inches internal diameter and about 10 inches long. A heater made out of a nichrome wire was introduced into the tube and the ends were taken out and connected to power supply through a wattmeter and a dimmerstat. The preheater tube was insulated from outside by a paste of magnesia-asbestos, about 1 inch thick.

Immediately above the preheater was the fluidization column, a 1.5 inches internal diameter and 15 inches long M.S. pipe; a stainless steel wire screen (80 mesh) rested between the fludization column and the preheater. About 9 inch portion of the fluidization column was surrounded by a cooling jacket. The jacket consisted of a 9 inch long G.I. pipe with 3 inch internal diameter. It was welded concentrically to the fluidization column through two flanges.

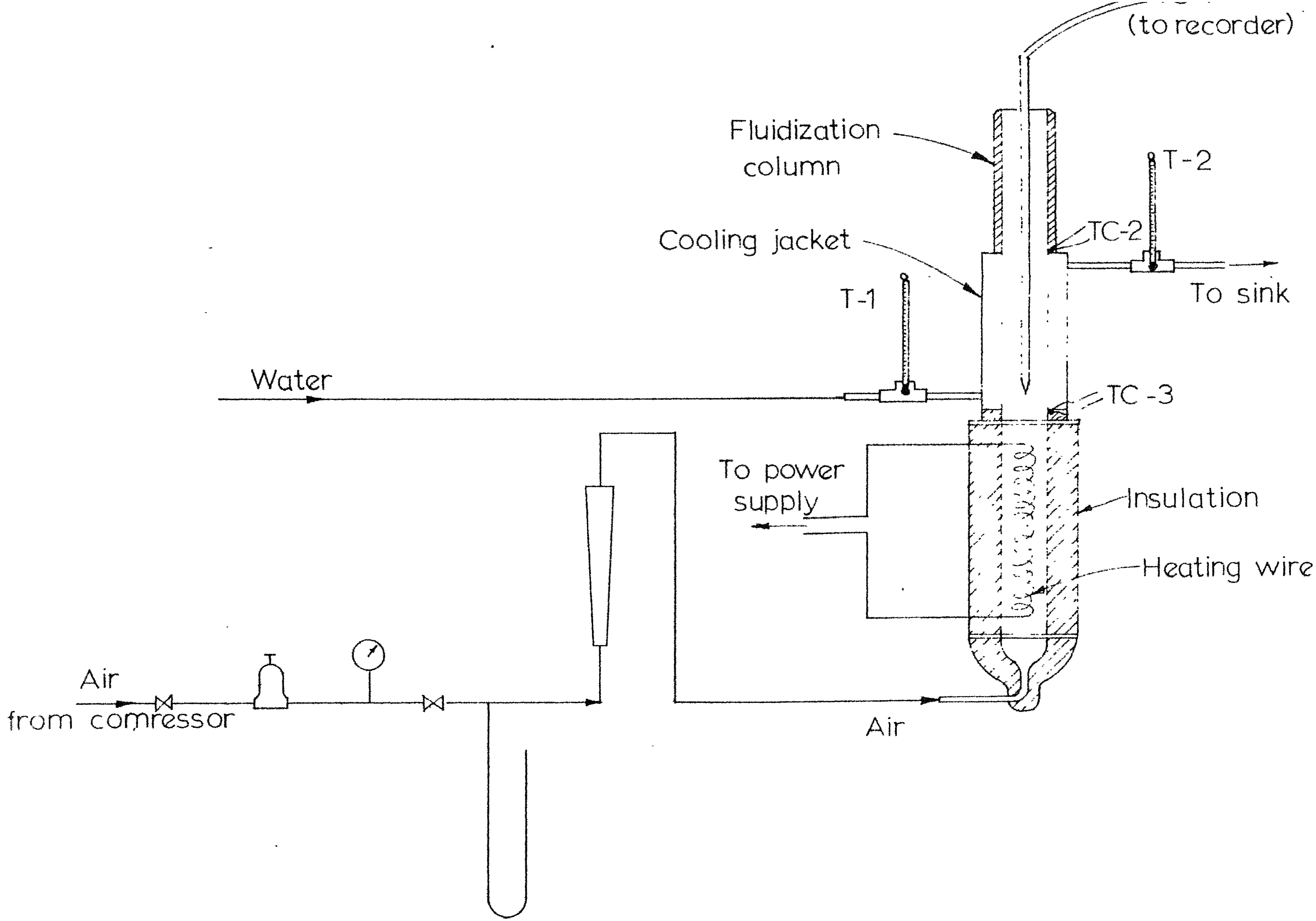


Fig. 1 - Flow sheet for heat transfer studies without chemical reaction.

The jacket was provided with inlet and outlet for water to circulate. The inlet and outlet temperatures were measured with precision (0.2°C) thermometers. The water flowrate was measured with a measuring cylinder.

Two thermocouples - one below the jacket and the other above it - were attached to the wall of fluidization column to give the wall temperature at the two positions. A third thermocouple which could be moved horizontally as well as vertically was used to measure the bed temperature at various positions. The thermocouples were connected to a " Honeywell, six point, temperature recorder ".

The portion of the fluidization column not covered by the jacket and the conical portion below the preheater were insulated with magnesia-asbestos paste.

Procedure for the Experimental Run :

The air supply was started at the predetermined rate. The water flowrate was maintained constant. The electric supply was switched on and the voltage adjusted. After about 15 minutes, when the two thermometers and the three thermocouples showed steady state, the whole set of observations i.e. water flow rate, the inlet and outlet water temperatures, rotameter reading and manometer reading, voltage and wattage supplied and the three thermocouple readings, was noted.

Using the above procedure, number of sets of observation for empty tube as well as with the bed were taken.

HEAT TRANSFER WITH CHEMICAL REACTION :

Hydrochloric Acid Gas Generation :

To carry out the chemical reaction, steady streams of HCl gas and ammonia gas were needed. The latter was obtained from the

compressed and liquified gas-cylinder, whereas the former was generated in the laboratory. For this purpose Brauer's [4] procedure was adopted which consisted of dropping concentrated hydrochloric acid in concentrated sulphuric acid when a stream of dry HCl gas is generated.

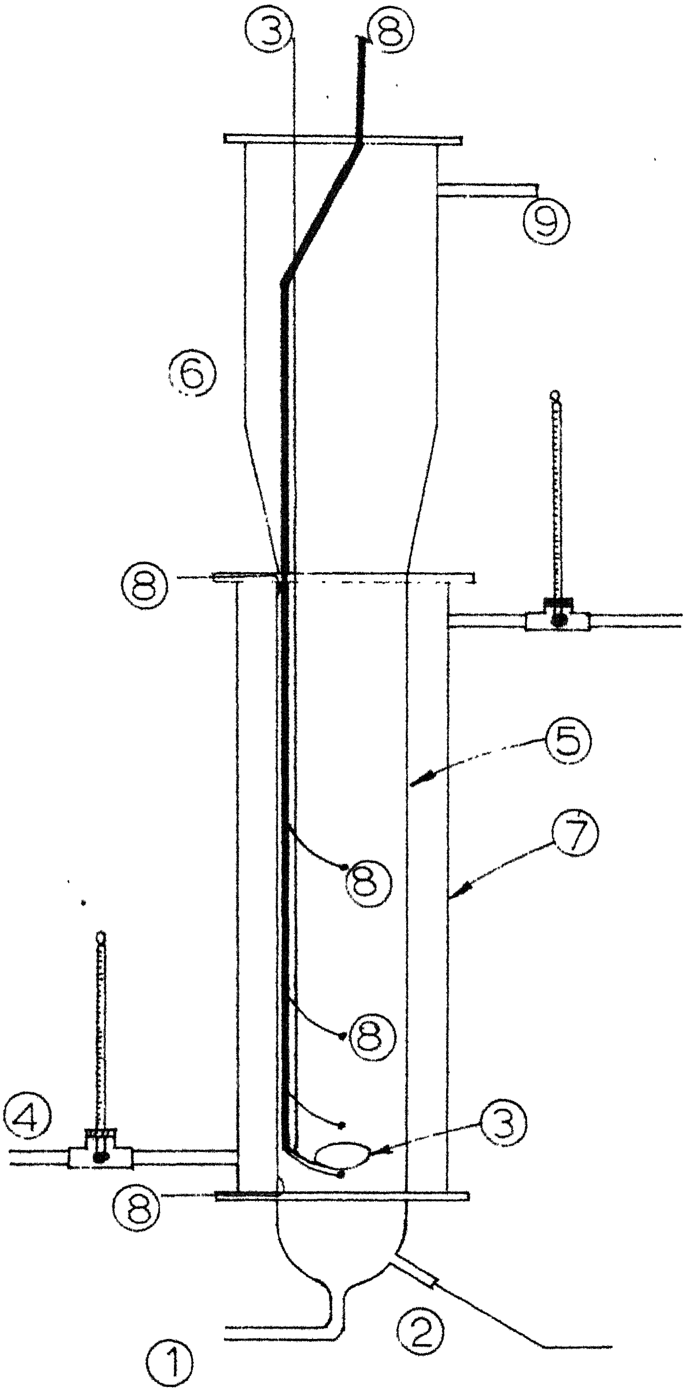
Experimental Set-up :

The fluidization set up (Fig. 2-a) consisted of a fluidization column - one 16 inch long, 3 inch internal diameter M.S. pipe - surrounded by a cooling jacket, 16 inch long, 4 inch internal diameter G.I. pipe, having one inlet and one outlet for cooling water. The flowrate of water was metered by a rotameter and the two thermometers were used to measure inlet and outlet temperature. Above the fluidization column, there was dis-engagement section. This section had one outlet for the exit of air and ammonium chloride particles. This was led to a trap consisting of a reservoir of water and the exit end of the tube dipped in water.

Two thermocouples were attached to the fluidization column, one near the bottom and the other near the top, to find the wall temperature. There were three thermocouples, at different axial positions of the bed to record the temperatures. Another thermocouple was kept sufficiently above the bed to record the temperature of the exit air. All the six thermocouples were led to a " Honeywell temperature recorder ".

(-72 + 80) mesh sand particles constituted the bed.

Hydrochloric acid gas generation unit consisted of a graduated separating funnel containing concentrated hydrochloric acid, one stainless steel, needle valve, one rotameter and a conical flask containing concentrated sulphuric acid. HCl gas stream was fed below the bottom of the fluidization column.



1 Inlet for air

2 Inlet for HCl gas

3 Ammonia feed system with a distributor at the bottom

4 Inlet for cooling water

5 Fluidization column

6 Disengagement section

7 Cooling jacket

8 Thermocouples connected to recorder

9 Outlet for air & ammonia chloride

Fig. 2a.-Fluidization setup- heat transfer with chemical reaction.

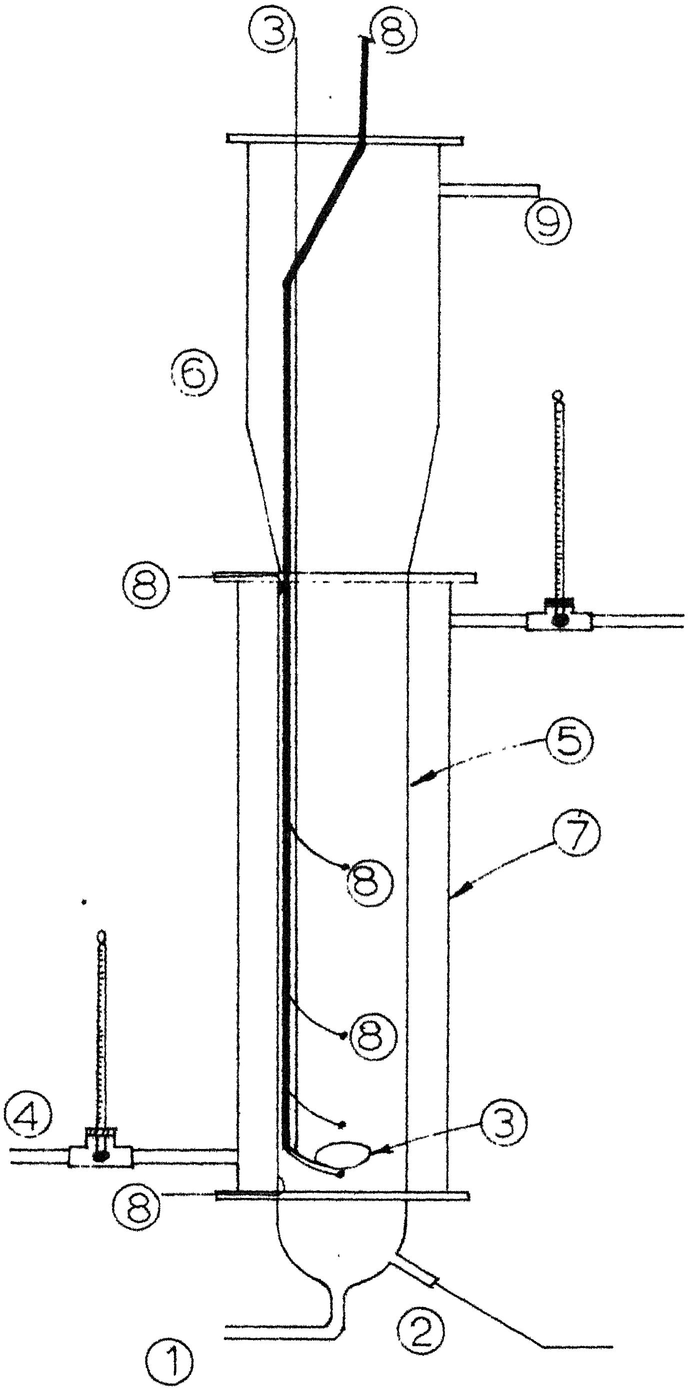


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Fig. 2a.-Fluidization setup- heat transfer with chemical reaction.

The flowsheet is given in Fig. 2(b).

Air supply was taken from a compressor in two parallel lines. Each line had a pressure regulator, pressure gauge, needle valve, mercury manometer, and rotameter. The air was fed at the bottom of the reactor. The supply of ammonia gas was taken from gas cylinder. Ammonia line also had one needle valve and one rotameter. The gas was fed into the reactor through a stainless steel 1/8" I.D. tube. The tube had the feed end near the top of the fluidization setup. The exit end was near the bottom the same and it consisted of a coil with one helix with holes over its length so that ammonia gas came out from number of holes.

Procedure :

The air supply was started at a pre-determined rate. After the steady state was shown by different thermocouples and thermometers, it was stopped momentarily, and hydrochloric acid supply was started at a predetermined rate to generate HCl gas stream. The air supply was resumed and ammonia gas supply was started. After about 15 minutes of reaction, hydrochloric acid supply was stopped. By that time an almost steady state had reached. In the mean time, flow rate of hydrochloric acid was checked to maintain a constant average supply. Before the supply of the acid was stopped, the complete set of observations i.e. flow meter readings for water, air, ammonia and hydrochloric acid, temperatures of inlet and outlet water and recorder readings,was noted.

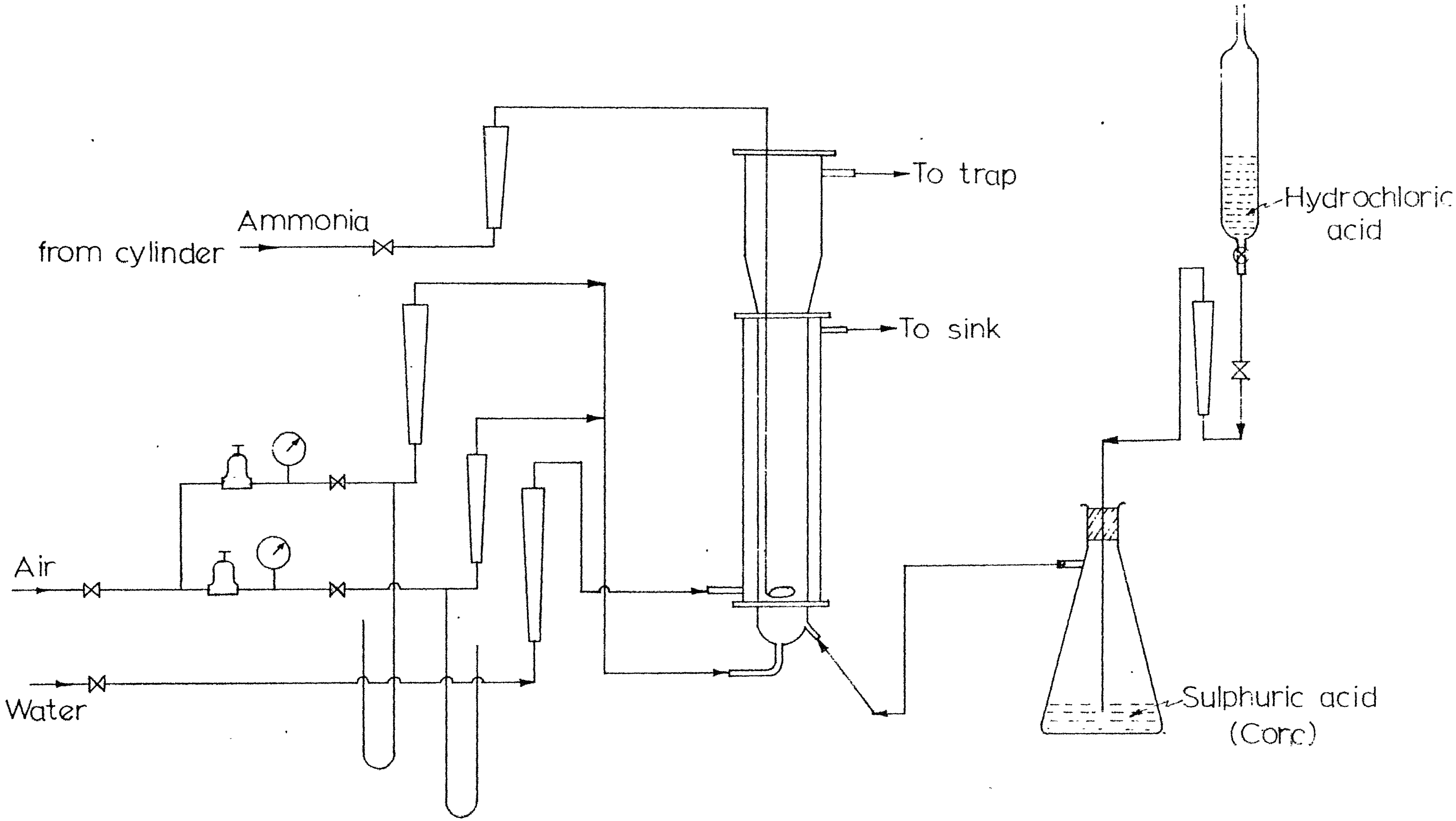


Fig. 2b - Flow sheet for chemical reaction carried in a fluidized bed.

CHAPTER 4

DISCUSSION OF RESULTS

HEAT TRANSFER WITHOUT CHEMICAL REACTION :

Table 1 gives the values of heat transfer coefficient for the corresponding mass velocity. In Fig. 3, the values have been plotted.

To explain the trend of the experimental data, let us consider the correlations proposed by Leva and coworkers (13) and Leva (12), namely,

$$h = 0.64 \frac{k}{\mu} G_f \eta \qquad \ldots (3)$$

and

$$h = 0.16 \frac{k}{D_p} \left[\frac{C_s \rho_s D_p^{1.5} g_c^{0.5}}{k} \right]^{0.4} \times \left[\frac{G_f D_p \eta}{\mu R} \right]^{0.36} \qquad \ldots (2)$$

The values of heat transfer coefficient given by the above equations have been plotted in the same figure. We find that at lower values of mass velocity, the slope of experimental curve is very close to 1, and for higher values, the slope is near to 0.36. That is, at lower values of mass velocity, the dependence is much more than at higher values.

The variation in dependence may be explained on the basis of quality of fluidization and bed homogeneity. At lower values of reduced mass velocity, the fluidization is quite smooth; but, at higher values, quality of fluidization deteriorates. This is further supported by the fact that at still higher values, a maximum is encountered as reported by Massimillah & coworkers (12).

TABLE 1

HEAT TRANSFER COEFFICIENT MASS VELOCITY RELATIONSHIP WITHOUT CHEMICAL REACTION

Bed height = 4.45 cm. for runs 101 - 105) (-52 + 72)
= 8.3 cm. for runs 106 - 110) mesh particles

S. No.	Run No.	Mass Velocity G_F, lbs/hr.sq.ft.	Ht. Tr. Coeff. h, BTUs/hr.sq.ft.
1	101	304	12.4
2	102	415.2	27.21
3	103	527.6	36.44
4	104	632.4	39.12
5	105	758.8	38.78
6	106	357.7	16.87
7	107	416.1	25.45
8	108	527.6	33.13
9	109	642.6	34.29
10	110	746.1	36.98

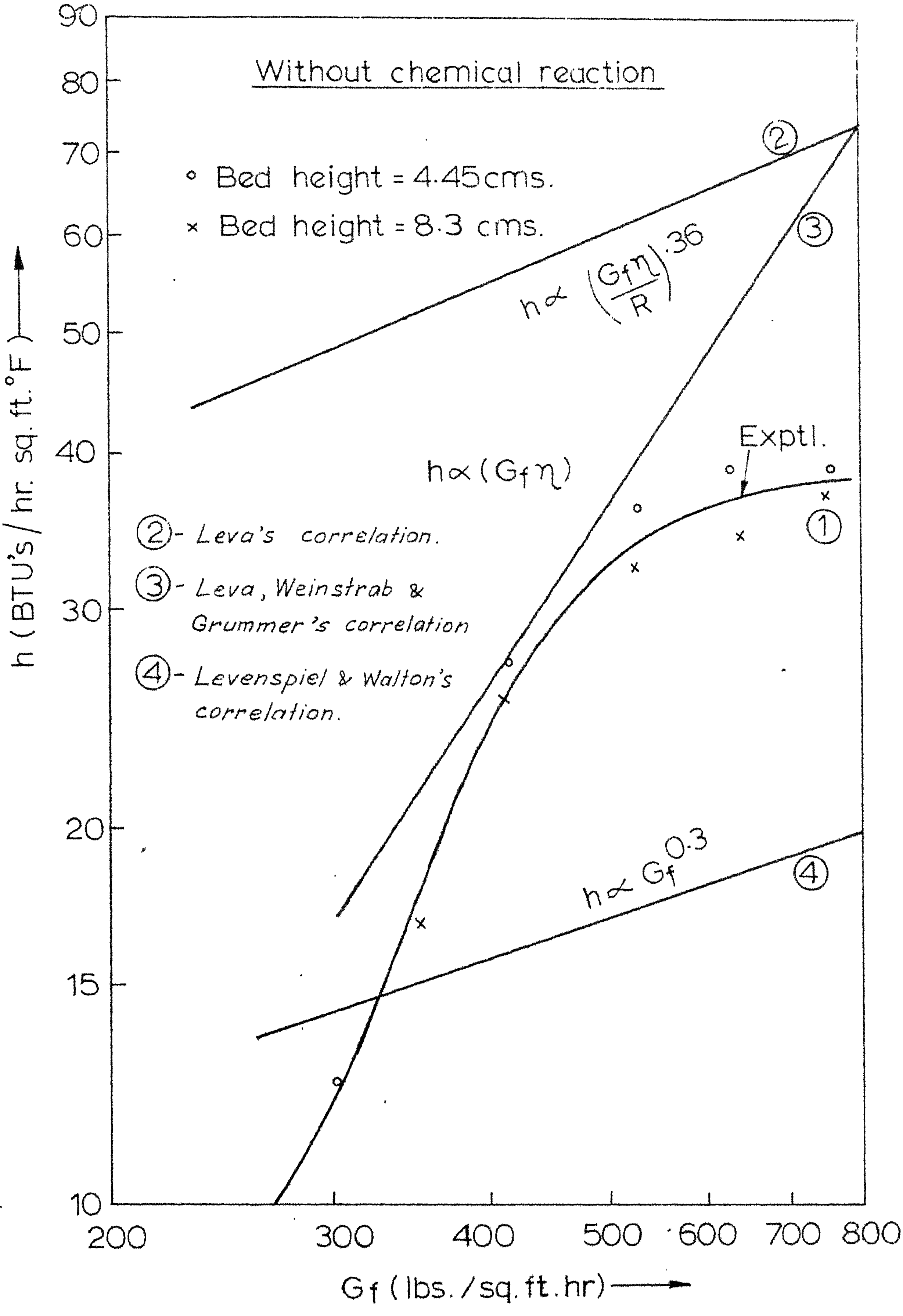


Fig. 3 - Heat transfer coefficient mass velocity relationship.

HEAT TRANSFER WITH CHEMICAL REACTION :

Heat Transfer Coefficient and Mass Transfer velocity Relatively :

In Table 2 and Fig. 4, the values of heat transfer coefficient against those of mass velocity have been given. Also plotted are the values given by the correlations proposed by different workers. The experimental values are quite high as compared to those obtained experimentally without chemical reaction, although trend remains the same.

This increase in values might have been due to (i) decrease in sand particle size (.0076" in the chemical reaction and .01" without the reaction), (ii) improvement in the quality of fluidization due to the presence of ammonia gas distributor and (iii) formation of ammonium chloride - additional solid particles.

Effect of Heat Liberated on Heat Transfer Coefficient :

In Table 3 have been given the values of heat transfer coefficient for different runs and the corresponding quantities of heat liberated. It is evident that there is no definite relationship between the heat liberated, QGEN, and the heat transfer coefficient h. In the present set of experiments, it was found that $h \propto G_f^{0.24}$. So, to take into account the mass velocity affect, $h/G_f^{0.24}$ has been also given in the Table and plotted in the Figure 5. Again, there is not of much dependence evident. So, it is concluded that heat transfer coefficient is independent of heat liberated.

Note : There were two different set ups and only one type of study was carried in each of them.

TABLE 2

HEAT TRANSFER COEFFICIENT
MASS VELOCITY RELATIONSHIP WITH CHEMICAL REACTION

Bed height = 6 - 8 cm. (-72 + 80) Mesh particles.

S. No.	Run No.	G_f	h
1	17	106.8	23.3
2	20	139.1	36.1
3	10	144.1	37.4
4	28	144.2	29.2
5	14	166.5	39.9
6	11	172.4	40.9
7	12	193.8	42.3
8	21	221.6	42.8
9	23	227.1	44.2
10	26	229.5	43.6
11	22	230.6	47.2
12	15	247.6	47.1
13	16	267.1	49.2
14	21	269.1	56.7

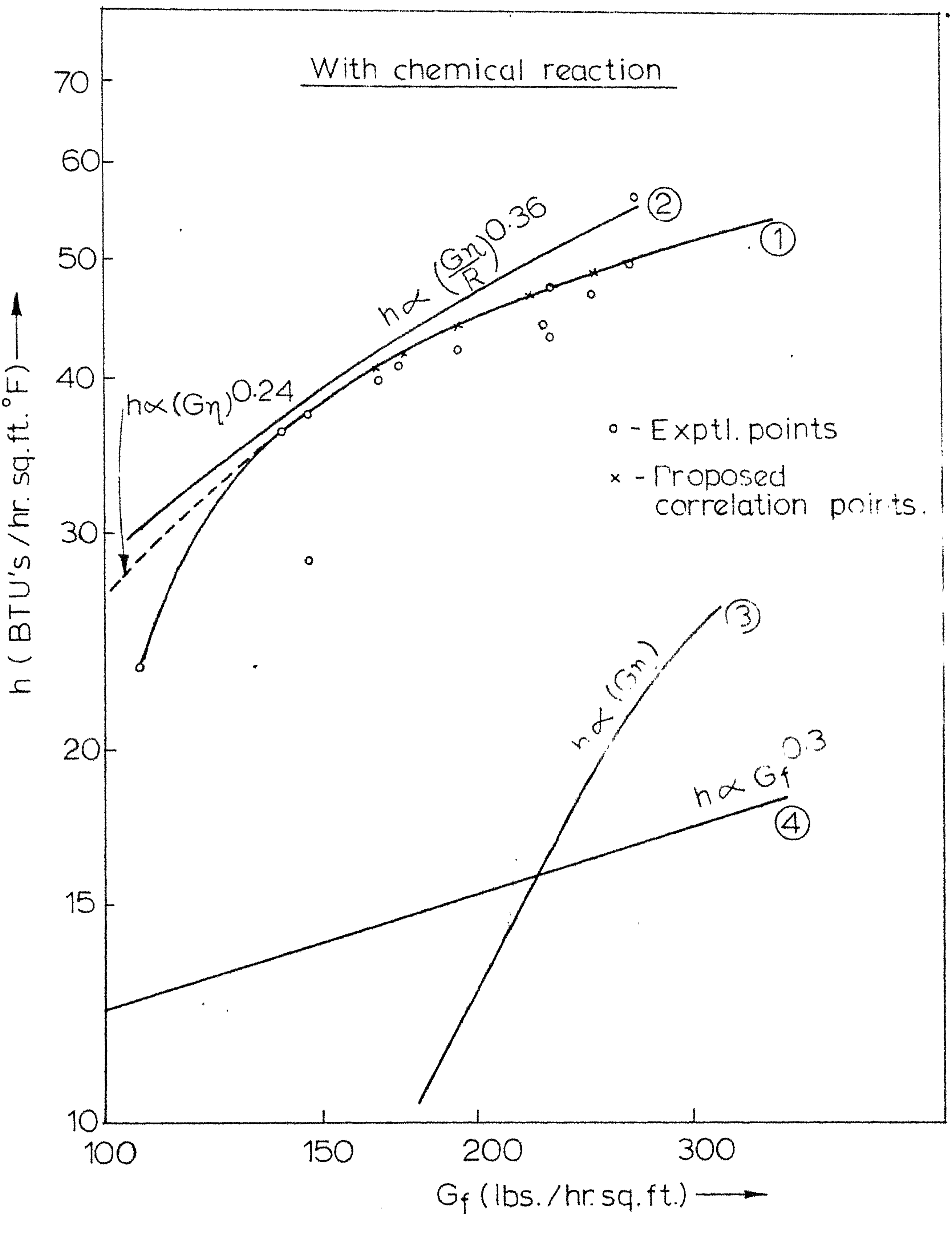


Fig.4 - Heat transfer coefficient mass velocity relationship.

TABLE 3

HEAT TRANSFER COEFFICIENT DEPENDENCE ON HEAT LIBERATED

S. No.	Run No.	Mass Vel. G_f lbs/hr.ft.	Heat Lib. QGEN BTUs/hr.	Heat Tr. Coeff. h BTUs/hr.sq.ft.°F	$\frac{h}{G_f^{0.2?}}$
1	28	144.2	1237.4	29.2	8.84
2	20	139.1	1343.0	36.1	11.05
3	14	166.5	1368.4	39.9	11.7
4	11	172.4	1990.4	40.9	11.86
5	10	144.1	2112.3	37.4	11.13
6	22	130.6	1648.2	47.2	12.9
7	23	227.1	1063.4	44.2	12.05
8	15	247.6	1118.4	47.1	12.51
9	29	221.6	1242.8	42.8	11.7
10	26	229.5	1742.2	43.6	11.84

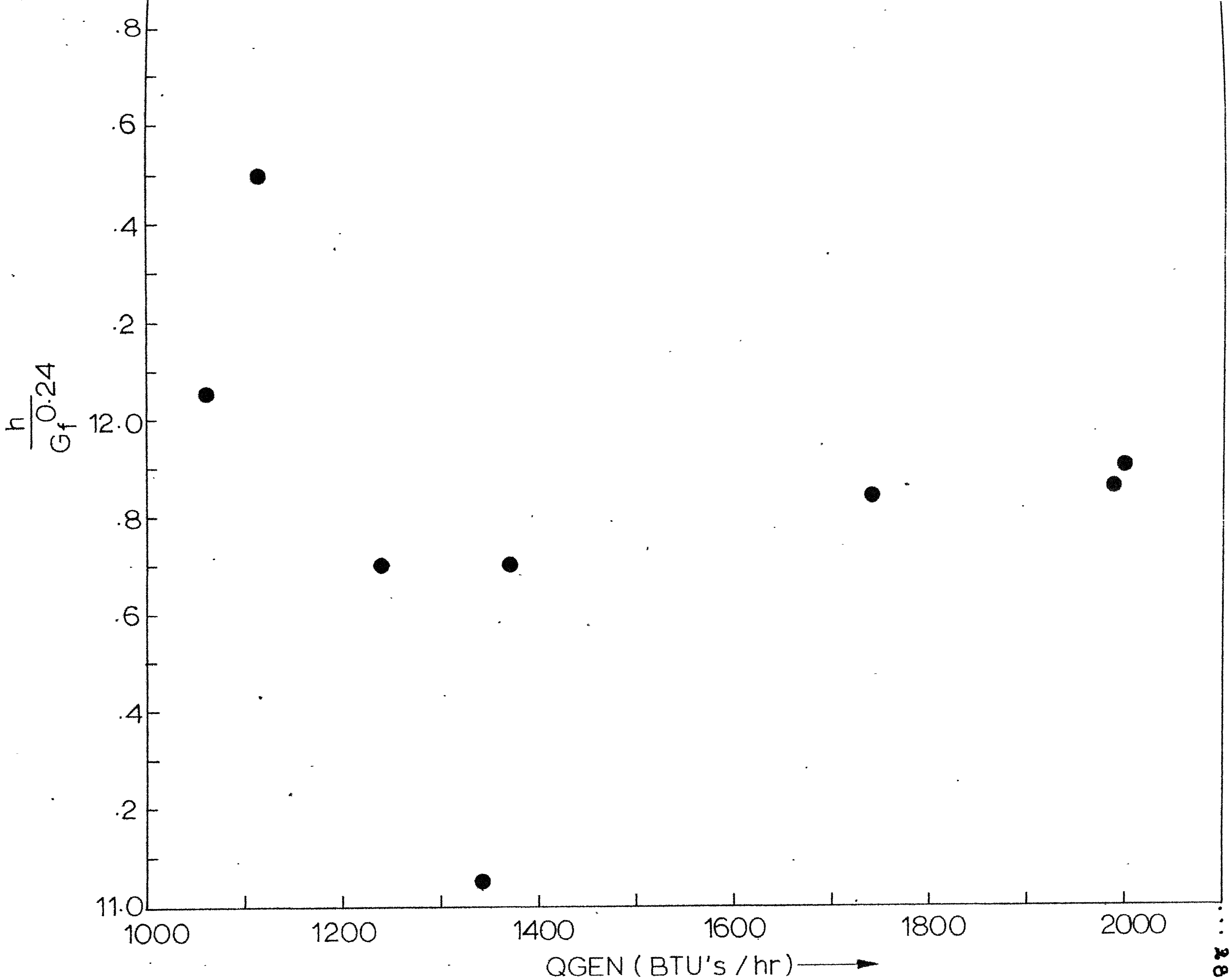


Fig. 5-Heat Tr. Coefficient Dependence on Heat Liberated.

Bed Temperature as a Function of Amount of Heat Liberated at Constant Mass Velocity :

In Table 4, the difference between the average bed temperature and the wall, $(T\ bed - T\ wall)_{av.}$, have been tabulated. The temperature profiles have been plotted in Figures 6a, 6b & 6c. As expected, the temperature difference increases with increase in the magnitude of QGEN (Fig. 6a & 6b). Again, keeping the magnitude of QGEN constant average bed temperature decreases with increase in mass velocity (Fig. 6c) runs 16 & 17. That is, the mass velocity has opposite effects on the two factors namely bed temperature and the heat transfer coefficient.

Again, the limiting mass velocity that can be usefully employed is the value where only ammonium chloride particles are carried away and not the sand particles. At the same time, the flow rates of reactant streams have to be such that ammonium chloride does not sublime because of rise in temperature of the bed because of heat liberated.

Ratio of Exit Air Temperature to that of Bed temperature :

It was expected that exit temperature of air will be nearly same as that of the bed (experiments carried without chemical reaction, Heertjes[10]). But, in the present investigation, $\theta = \frac{(T_{air} - T_{wall})}{(T_{bed} - T_{wall})}$ was found to be strongly dependent upon the mass velocity of air, and, its value varied from about 0.5 to about 2, as given in Table 4 and Fig. 7. One explaination for this observation is that height of the bed was too small (6-8 cm.) to attain the equilibrium between the two temperatures. Secondly, it suggests a mechanism for heat transfer also. That is, at lower mass velocities, heat liberated is mainly 'picked up ' by the solid particles and then they heat the air in

TABLE 4

(a) BED TEMPERATURE DEPENDENCE ON HEAT LIBERATED

(b) DIMENSIONLESS AIR TEMPERATURE DEPENDENCE ON MASS VELOCITY

S. No.	Run No.	G_f	QGEN	$T_{bed} - T_{wall}$ °C	$\theta = \frac{T_{air} - T_{wall}}{T_{bed} - T_{wall}}$
1	22	230.6	648.4	15.1	2.067
2	23	227.1	1063.4	34.8	0.509
3	29	221.6	1242.8	36.0	1.31
4	26	229.5	1742.2	50.7	0.463
5	28	144.2	1237.4	47.0	0.446
6	20	139.9	1343.0	46.5	1.054
7	10	144.1	2112.3	68.5	1.165
8	11	172.4	1990.4	48.2	0.998
9	12	193.8	1997.9	54.0	1.575
10	14	166.5	1368.4	40.0	1.125
11	15	247.6	1118.4	29.3	1.64
12	16	247.1	1203.2	33.2	1.210
13	17	106.8	1207.4	66.5	0.575
14	21	269.7	1287.9	35.0	0.9

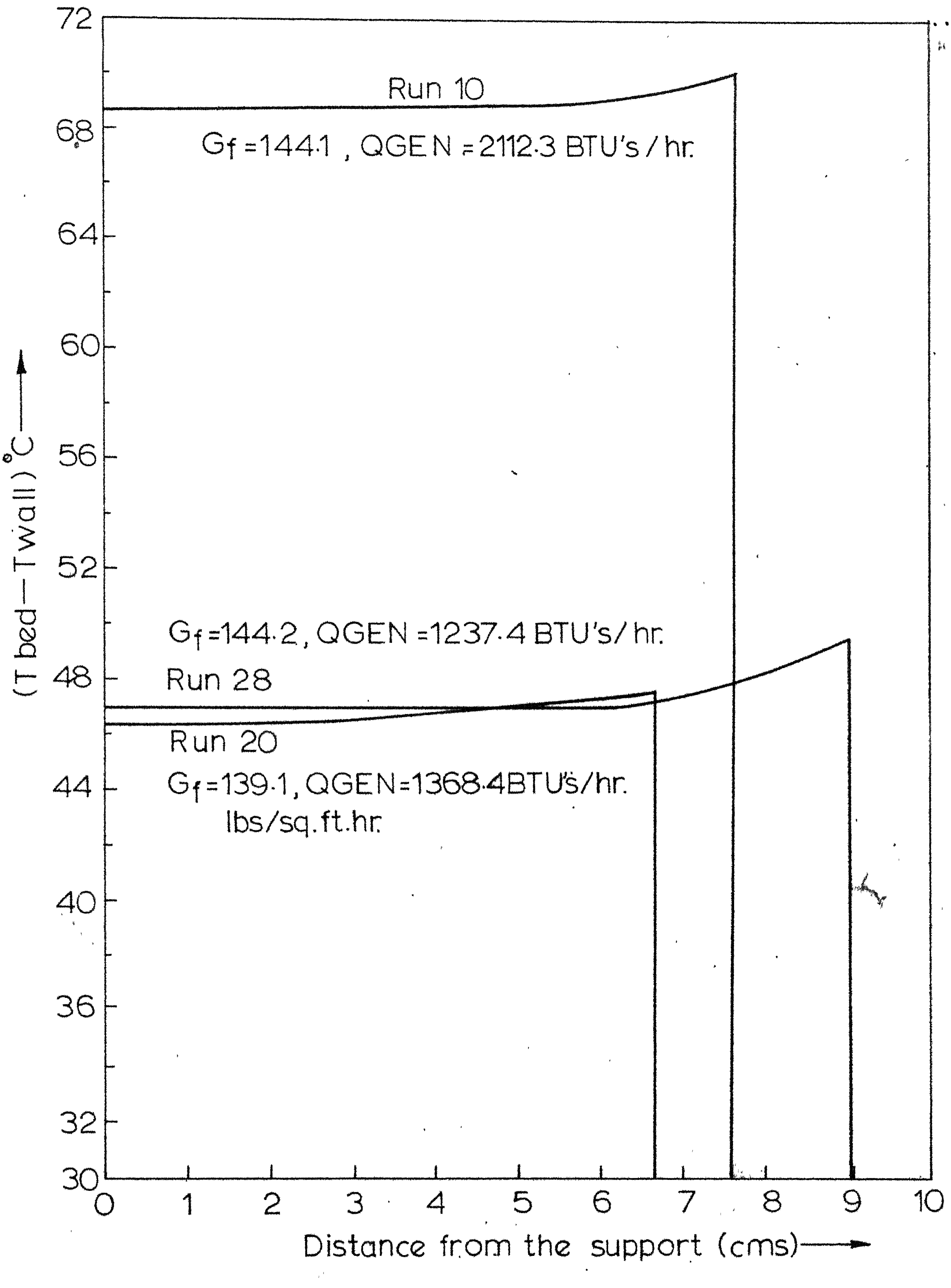


Fig.6a - Bed temperature and heat liberated relationship.

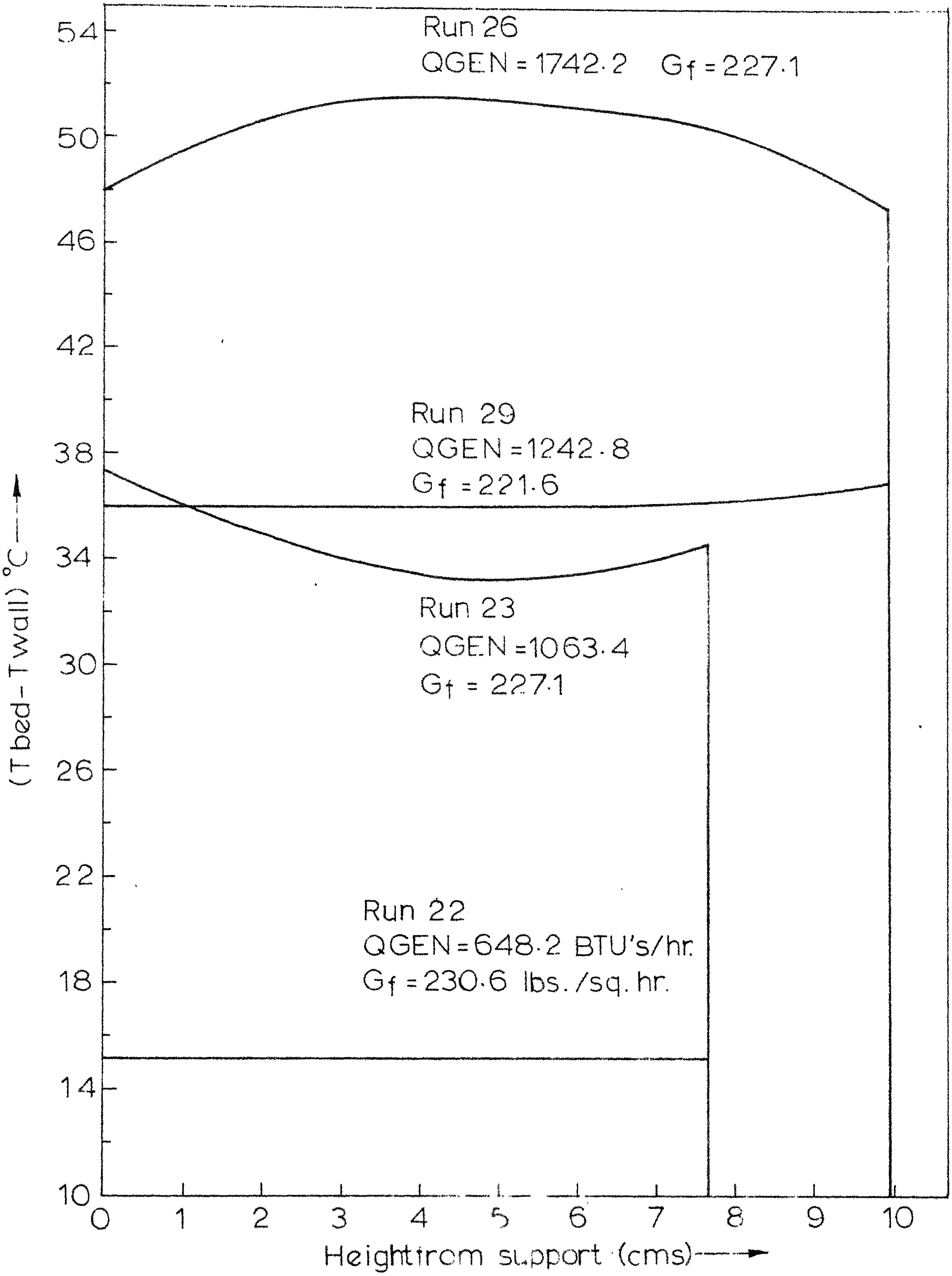
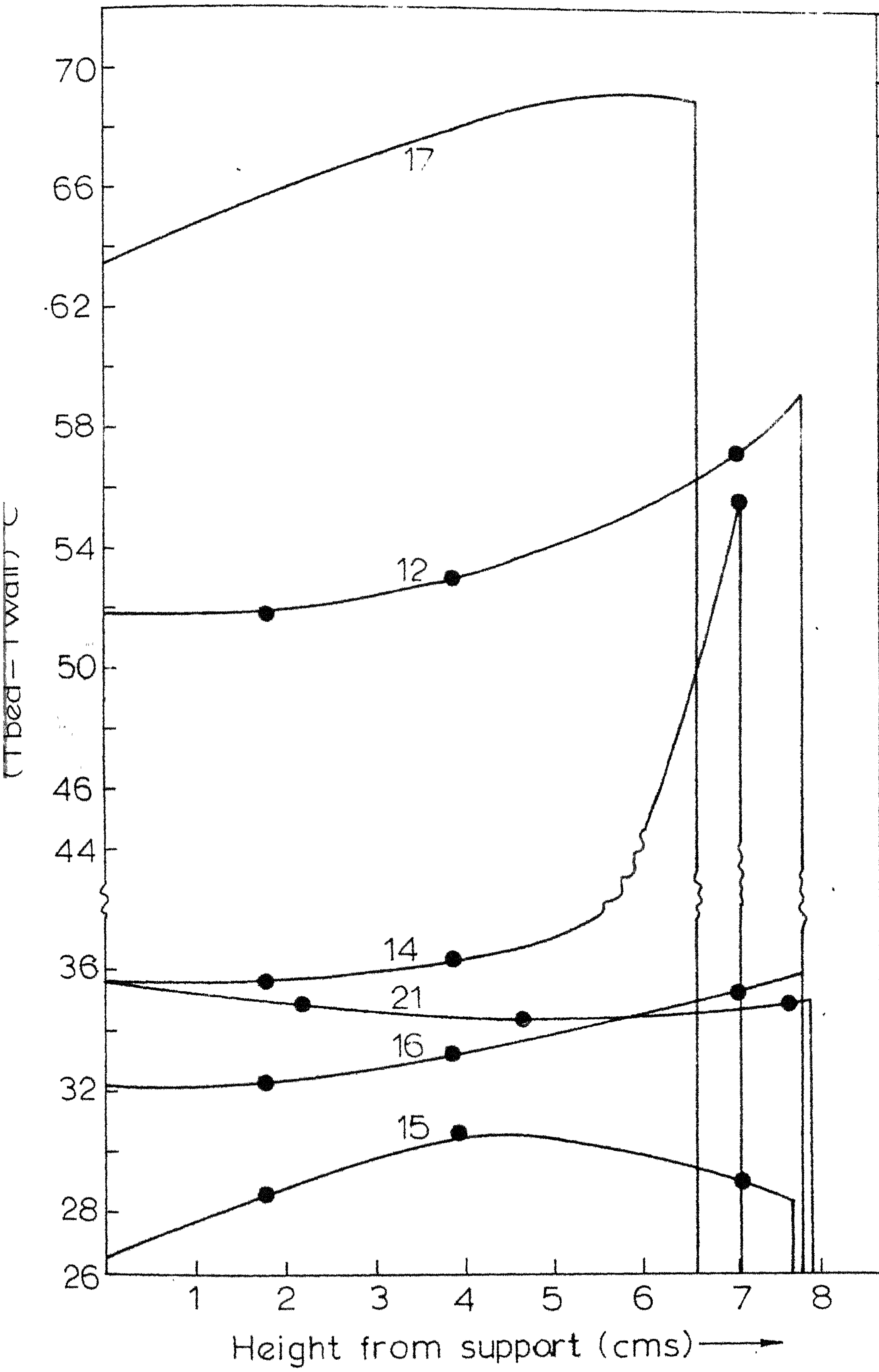


Fig. 6b - Bed temperature and heat liberated relationship.



RUN	G_f	QGEN
17	106.8	1207.4
14	166.5	1368.5
11	172.4	1990.4
12	193.8	1997.9
15	247.6	1118.4
16	267.1	1203.2
21	269.7	1287.9

Fig.6c - Bed temperature and heat liberated relationship.

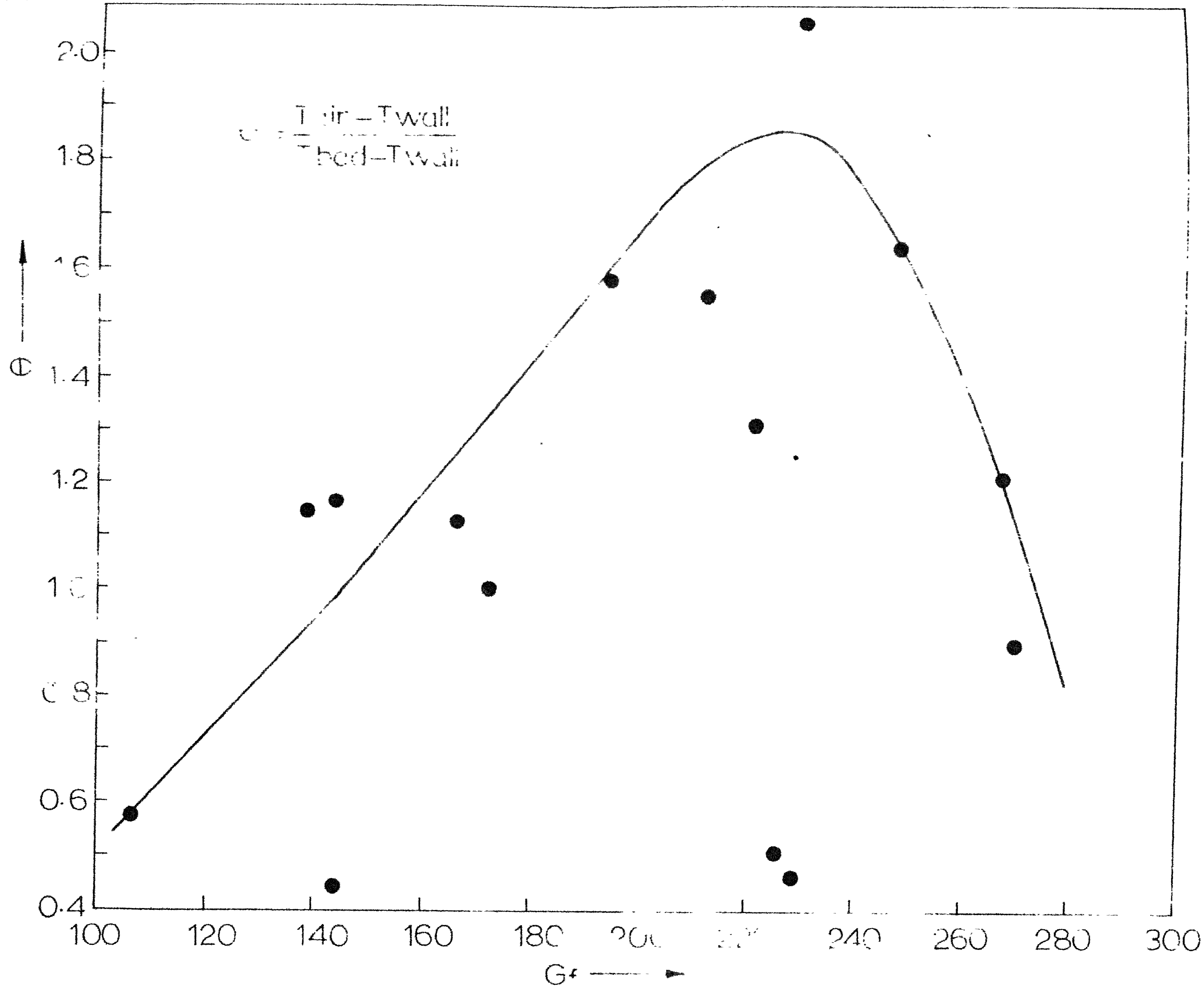


Fig 7

air

Dimensionless temperature and mass velocity relationship.

turn. That is why θ is less than 1 under such conditions. At sufficiently higher velocities-around the region where maximum is encountered - heat linerated is mainly 'picked up' by air because of increased voidage and heat transfer takes place from air to particles, resulting the value of θ to be greater than unity.

At still higher mass velocities, θ tends to be equal to unity, the reason being that mass velocities are so high that temperature rise for air is not so much. At intermediate values of mass velocity, θ is again around unity the explaination being that both the mechanisms, mentioned above, play a significant role.

It may be noted that when $\theta < 1$, the air temperature in the bed will still be less than that of outgoing ~~due~~ one. For $\theta > 1$, the opposite thing will hold good. The reason for these is that in both the cases, heat transfer is taking place upto the point they are in contact with each other.

CONCLUSIONS :

1. The magnitude of heat transfer coefficient increased with chemical reaction. It might have been due to (a) formation of additional solid particles (ammonium chloride), and (b) ammonia gas distributor resulting in better contact.
2. Dependence of magnitude of heat transfer coefficient on mass velocity at lower values of reduced mass velocity is higher than at higher values of the same ; $h \propto (G_f)^{0.24}$, for $\frac{G_f}{G_{mf}} > 2.25$
3. (a) Keeping mass velocity constant, the temperature of the bed increases in magnitude with the heat liberated Fig.6a,&6b.
   (b) For the same quantity of heat liberated, the bed temperatur decreases with increase in the mass velocity (Fig.6c, runs 16&1
4. The ratio of exit air temperature to that of the bed is a stron function of mass velocity. It shows a maximum. (Fig. 7).

APPENDIX

TABULATION OF DATA AND METHODS OF CALCULATIONS

HEAT TRANSFER WITHOUT CHEMICAL REACTION :

Heating wire in preheater :

Length = 21 ft. approx.; Gauge = 21 BWG (0.035"); Total resistance = 22 Ohms.

Fluidization Column :

Length = 15.2 inches ; I.D. = 1.51 inch.

General :

Room temperature of water = 20°C (approx.)

Room temperature of air = 21°C (approx.)

Atmospheric pressure = 740 mm. (Mercury) approx.

For air, μ = .0207 c.p. ; k = .01743 BTUs/hr. ft.°F.

Mercury manometer difference = 5cm. approx.

Rotameter calibration chart was available; the calibration was at 1 atm. pressure and 70°F.

Vol. flow rate from the calibration chart was directly used to find the mass velocity.

EMPTY TUBE :

This part of experiments was necessitated by the fact that in the process of preheating the air, some of the heat passed along the walls of the preheater by conduction. This amount depended upon (i) wattage, and (ii) flow rate of air. In Table 5, the data, observed values as well as the calculated ones, have been tabulated.

TABLE 5

FRACTION OF HEAT SUPPLIED-CARRIED THROUGH CONDUCTION-DEPENDENCE ON AIR FLOWRATE

Serial No.	Flow Rate		Temperature Rise		Energy		$x_{cond.}$
	Air, lit/min.	Water, gms/min	Air, °C.	Water, °C	Supplied watts	Passing by cond. BTUs/hr	
1	23.8	154.0	54	4.8	160	128	.235
2	41.3	210.0	95	3.4	164	61.6	.113
3	51.0	212.0	85	3.4	160	70.5	.129
4	59.4	200.0	74	3.3	160	66.0	.121
5	68.0	192.0	63	2.8	160	92.3	.169
6	15.7	218.0	138	6.2	320	35.7	.135
7	23.8	291.0	179	4.3	320	97.5	.089
8	31.7	408.0	175	2.9	326	79.0	.071
9	41.3	532.0	180	2.2	320	52.6	.048
10	50.3	600.0	161	2.0	312	72.4	.068
11	58.2	584.0	157	2.0	326	61.0	.055
12	69.0	568.0	141	2.1	322	80.7	.073

...37

<u>Method of Calculations for Any Set of Observations :</u>

(i) Reynold's number, N_{Re}, w.r.t. tube diameter was found.

(ii) Depending upon the value of Reynold's number heat transfer coefficient for the empty tube was found (Perry pp.10-13 and 10 -14).

(iii) From the temperature conditions and heat transfer coefficient. QA, convective heat transfer to the wall was found.

(iv) QCON = QW-QA, gave the amount of heat taken up by water because of conduction along preheater walls, where, QW = total amount of heat taken up by water.

<u>Sample Observations and Calculations :</u>

For observation No. 7 in Table 5, corresponding to flow rate of 23.8 liters/min.,

h_{air} = 1.24 BTUs/hr. ft.°F. h_{air} = heat transfer coefficient for air

$$QA = h_{air} \times (T_{air} - T_{wall}) \times \text{Internal surface area of the tube}$$

$$= 200.4 \text{ BTUs/hr. (not given in table)}$$

$$QW = \frac{291.8 \times 4.3 \times 1.8 \times 60}{252} = 297.9 \text{ BTUs/hr.}$$

$$QCON = 297.9 - 200.4 = 97.5 \text{ BTUs/hr.}$$

$$x_{cond.} = \frac{\text{Amount of heat taken up by water through conduction}}{\text{Total amount of heat supplied}}$$

$$= \frac{97.5}{320 \times 3.412} = .089$$

x_{cond} values were plotted as a function of flow rate and wattage supplied as parameter. They were made use in subsequent calculations.

FLUIDIZED BED :

Bed :

Height = 4.45 cms. for runs 101 - 105

= 8.3 cm. for runs 106-110

Diameter of particles = .01" (-52 + 72) B.S.mesh

Min. fludization mass velocity = 88.5 lbs/hr. sq.ft.

In Table 6, the data, observed values as well as the calculated ones have been tabulated.

Method of Calculations :

(i) Reynold's number w.r.t. tube diameter was found and hence the value of heat transfer coefficient.

(ii) Convective heat transfer, QA, by air was found as before.

(iii) QW - (QA + QCON) = QF, heat transferred by the fluidized bed was found, taking the value QCON (from $x_{cond.}$) from the blank runs, conducted previously.

(iv) Heat transfer coefficient, h, was found by making use of the integrated temperature difference and bed expansion ratio.

Sample Calculations:

For run no. 101,

Corresponding to a flow rate of air of 23.8 lit/min and wattage of 160, $x_{cond.}$

G_f = mass velocity of air

$$= \frac{23.8 \times 60}{(3.048)^3} \times \frac{29 \times 492}{359 \times (460 + 70)} \Big/ \frac{\pi}{4} \times \left(\frac{1.51}{12}\right)^2$$

= 304 lbs/hr. sq.ft.

Reduced mass velocity $= \frac{304}{88.5} = 3.431$

TABLE 6

HEAT TRANSFER COEFFICIENT MASS VELOCITY RELATIONSHIP

Serial No	Runs no.	Flow rate		$T_{bed}-T_{wal}$ °C	Temp.rise Water °C	Energy supplied watts	$x_{cond.}$	Bed expansion ratio	G_f	h
		Air, lit/min	water gms/min							
1	101	23.8	415	38	2.0	160	.190	1.33	304	12.54
2	102	32.5	420	36	2.6	165	.150	1.45	415.2	27.21
3	103	41.3	424	33	3.0	164	.125	1.55	527.6	36.44
4	104	49.5	424	31	3.0	164	.122	1.64	632.4	39.12
5	105	59.4	425	31	3.2	164	.138	1.74	758.8	38.78
6	106	28.0	426	40	2.8	318	.079	1.38	357.7	16.87
7	107	32.1	452	40	3.6	317	.071	1.44	416.1	25.45
8	108	41.3	449	40	4.6	316	.059	1.55	527.6	33.13
9	109	50.3	450	44	5.4	320	.061	1.65	642.6	34.29
10	110	58.2	450	41	5.8	320	.063	1.78	746.1	36.98

... 40

Corresponding bed expansion ratio = 1.33 (12)

Average temperature from integrated temperature difference = 38°C

$$h = \frac{QF}{\pi \times \frac{4.45}{30.48} \times \frac{1.51}{12} \times 1.33 \times 38° \times 1.8} = 12.54$$

WITH CHEMICAL REACTION

Total length of the fludization column = 16.2"

Diameter of the column = 3.156"

Room temperature of water = 34°C (approx.)

Room temperature of air = 36°C (approx.)

Rotameter calibration charts for air were available; the calibration was at 1 atm. and 70°F. Mass flow rate under the existing conditions was calculated and used in other calculations.

Particle diameter D_p = .0076" (-72 + 80 mesh)

Static bed height = 5.9 cm. (= 2.31") for runs 10 - 23

= 7.7 cm. (= 3.03") for runs 26 - 29

Min.fludization velocity = 60.5 lbs/hr. sq.ft.

METHOD OF CALCULATION :

In Tables 7 & 8, the data have been tabulated.

(i) QA was found as before

(ii) QW was found from the flowrate and temperature rize of water data.

(iii) Attributing (QW-QA) to fluidized bed, heat transfer coefficient was found making use of bed expansion ratio, integrated temperature difference, area of heat transfer.

(iv) Total heat liberated per unit time, QGEN, was found from the flow rate of hydrochloric acid, gms. of HCl gas given out per uniy volume of acid and heat of reaction.

(v) Mass velocity of air under existing conditions.

For rotameter 1, $\rho_1 = \frac{29 \times 273 \times (74 + PMAN1)}{22.414 \times 76\ (273 + room)}$ gm./lit.

ρ_1 = density of air passing through rotameter1.

PMAN1 = manometeric pressure difference in the mercury manometer for rotameter 1, cm.

Similarly ρ_2 is evaluated.

mass velocity, gms./sq.cm.mm. = G_{CG}

$$G_{CG} = \frac{\rho_{rot.} \left[V_1 \times \sqrt{\frac{\rho_1}{\rho_{rot}}} + V_2 \times \sqrt{\frac{\rho_2}{\rho_{rot}}} \right]}{\text{Area of cross section}} \qquad \text{(5-13 Perry)}$$

where ρ_{rot} = Density of air in the conditions under which rotameter was calibrated.

G_f = Mass velocity, lbs/hr sq.ft.

$$= \frac{G_{CG} \times (30.48)^2 \times 60}{453.6} = 123\ G_{CG}$$

Sample Calculations :

For run 10, (i) amount of heat liberated/hr.

$$= \frac{27.7 \times 60}{36.5} \times \frac{55.8}{200} \times \frac{41,900}{252}$$

$$= 2112.3 \text{ BTUs/hr.}$$

55 gms of dry HCl liberated per 200 c.c. of conc. HCl; 41,900 calories /gm. mole, is the heat of reaction; 1 BTU = 252 calories.

Rest of the calculations are as before.

TABLE 7

OBSERVATIONS WITH CHEMICAL REACTION

Serial No.	Run No.	Rotameter I(Air)		Rotameter(II)air		Flow rate		Temperature rise		Wall Temp. °C.
		Flowrate lit/min.	Hg.Mano. cm.	Flowrate lit/min	Hg.Mano cm.	Water gms/min.	HCl ccs/min	Water °C.	Air °C.	
1	10	54.1	12.0	00	0	420	27.7	9.8	80.0	49.5
2	11	60.3	15.5	00	0	550	26.1	5.8	58.6	42.7
3	12	64.2	15.0	8.00	1.5	550	26.2	6.8	45.2	44.8
4	14	59.0	12.5	00	0	420	32.3	5.7	86.5	43.5
5	15	63.8	16.0	24.00	3.0	600	26.4	3.3	48.0	39.0
6	16	56.4	13.0	39.00	5.0	610	28.4	4.4	40.2	40.8
7	17	38.8	6.0	00	0	525	28.5	3.9	38.3	39.2
8	20	50.0	10.0	00	0	600	31.7	4.4	53.0	42.0
9	21	63.8	18.0	30.00	4.0	725	30.4	4.5	32.8	44.7
10	22	63.3	16.5	15.50	5.0	500	15.3	2.4	31.2	37.8
11	23	61.2	17.0	15.0	5.0	460	25.1	5.1	40.3	40.3
12	26	63.0	17.5	15.5	5.0	640	40.7	6.8	49.0	49.0
13	28	50.0	9.0	00	0	620	23.1	4.1	38.0	38.0
14	29	62.0	17.0	15.5	5.0	585	23.2	5.3	40.0	40.0

TABLE 8

CALCULATIONS OF RESULTS - WITH CHEMICAL REACTION

Serial no.	Run No.	G_f, lbs/sq.ft hr.	A_{curve}	QGEN, BTUs/hr.	QW, BTUs/hr.	QA, BTUs/hr	H.T.Coef h, BTUs/hr s ft.°
1	17	106.8	303	1207.4	487.5	35.3	23.3
2	20	139.1	258	1343.0	628.6	53.1	36.1
3	10	144.1	389	2112.3	980.0	80.1	37.4
4	28	144.2	324	1237.4	608.0	20.4	29.2
5	14	166.5	212	1368.4	570.0	47.6	39.9
6	11	172.4	280.5	1990.4	759.5	51.0	40.9
7	12	193.8	305.0	1997.9	890.5	93.6	42.3
8	29	221.6	270.0	1242.8	696.4	51.8	42.8
9	23	227.1	197.0	1063.4	558.6	20.6	44.2
10	26	229.5	375.0	1742.2	1036.2	26.2	43.6
11	22	230.6	86.0	648.4	285.7	36.5	47.2
12	15	246.7	162.0	1148.4	528.6	57.4	47.1
13	16	267.1	194.0	1203.2	639.0	49.2	49.2
14	21	269.7	210.0	1287.9	776.8	40.2	56.7

QW = Heat taken away by water ; QA = Convective heat transferred by air

Note

(1) Upto run no. 12, HCl (I), was used

From run no. 13 to 26 , HCl (II) was used

For remaining runs, HCl (III) was used.

200 cc. of conc. HCl (I) gave 55.8 gm. of HCl gas

200 cc. of conc. HCl(II) gave 31 gm. of HCl gas

200 cc. of conc. HCl (III) gave 39.2 gm. of HCl gas.

(2) All the calculations were based, taking one hour as the unit time, although almost all the observations were on one minute as the unit time.

(3) Conversion factors were made use of whenever any change of units occured.

EMPIRICAL CORRELATIONS USED FOR COMPARISON:

Leva, Weinstraub & Grummer's Correlation (13)

$$h = 0.64 \frac{k}{\mu} G_f \eta$$

Levenspiel & Walton's Correlation (14)

$$\frac{h}{C_F G_f} = 0.6 \left(\frac{D_p G_f}{\mu} \right)^{-0.7}$$

Leva's Correlation (12)

$$\frac{h D_p}{k} = 0.16 \left(\frac{C_s \rho_s D_p^{1.5} g_c^{0.5}}{k} \right)^{0.4} \left(\frac{G_f D_p \eta}{\mu R} \right)^{0.36}$$

Dow and Jakob's Correlation [5]

$$\frac{h\ D_t}{k} = 0.55\left(\frac{D_t}{L_f}\right)^{0.65}\left(\frac{D_t}{D_p}\right)^{0.17}\left[\frac{(1-\epsilon)C_s\ \rho_s}{\epsilon\ C_F\ \rho_F}\right]^{0.25}\left[\frac{D_t G_f}{\mu}\right]$$

ϵ = Voidage

Van Heerden, Nobel & Krevelens Correlation [8]

$$\frac{Nu}{Pr^{0.5}} \times \left[\frac{\rho_{mf}}{\rho_F}\right]^{0.18}\left[\frac{C_F\ \rho_F}{C_s\ \rho_{mf}}\right]^{0.36} = 0.58\ (BRe)^{0.45}$$

$$Nu = \frac{h\ D_p}{k}\ , \quad Pr = \frac{C_F\ \mu}{k}$$

B = Generalised shape factor , $Re = \frac{D_p\ G_f}{\mu}$

Toomey & Johnstone's Correlation [18]

$$\frac{h\ D_p}{k} = 3.75\left[\frac{D_p\ U_{mf}\ \rho_F}{\mu}\ .\ \log\ \frac{U_f}{U_{mf}}\right]^{0.47}$$

In table 9 the conditions mentioned by various workers have beem in listed.

TABLE - 9

Conditions mentioned by various workers

Ref.	Solids	Voidage Range	Absolute Density lb /c.ft.	Particle size range μ	Vessel Dia. in.	Bed Height in.	Fluids	Mass velocity range lb/hr.sq.ft
5	Aerocat, coke, Iron Powder.	52 - 69	121 - 466	110- 171	2.06 & 3.07	2 - 13	Air	50- 300
8	Carborundum, Iron Oxide, Coke, Lead fly ash.	Dense phase	37.5-694	80 -650	3.4	16	Air, CH_4, CO_2, town gas, H_2 & N_2 mixture	44 - 779
12	-	-	-	38 - 915	2.0-4.73	-	-	20- 1500
13	Sand, Iron catalyst, silica gel.	35 - 75	80 - 500	39 - 455	2.0 & 4.0	12 - 25	Air, CO_2 He, N_2	147 -1095
14	Glass beads, Catalyst, coal -	41.7 - 86.2	63.6-180	76-4430	4	10-30	Air	79 - 4350
18	Glass beads.	Dense Phase	167- 179	55- 850	4.73	13.2-246	Air	23.7-1542

... 47

LIST OF SYMBOLS

A_{curve} = Area under the curve, integrated temperature difference-bed height.

C_s = Specific heat of solid particles, BTUs/lb °F.

C_F = Specific heat of the fluid, BTUs/lb°F.

D_p = Particle diameter, inches.

D_t = Tube diameter inches.

g_c = Acceleration due to gravity, ft/sec^2

G_f = Mass velocity, lbs/hr. sq.ft.

h_{air} = Heat transfer coefficient for empty tube, BTUs/hr.sq.ft.°F.

h = Heat transfer coefficient for fluidized bed BTUS/hr.sqft.°F

k = Thermal conductivity of the gas, BTUs/hr.ft.°F.

L_f = Bed height, ft.

QGEN = Heat liberated, BTUs/hr.

QCON = Amount of heat passing by conduction, BTUs/hr.

QA = Heat transfer by convection, empty tube, BTUs/hr.

QW = Heat taken up by water, BTUs/hr.

R = Bed expansion ratio.

T_{air} = Exit temperature of air, °C.

T_{bed} = Bed temperature °C.

T_{wall} = Average wall temperature, °C.

U_f = Fluidization velocity of air ft/sec.

U_{mf} = Minimum fluidization velocity, of air ft/sec.

η = Fluidization efficiency

μ = Coefficient of viscosity, c. poise.

ρ_s = Density of solid particles, lbs/cu.ft.

ρ_m = Density of quiescent bed lbs/cu.ft.

ρ_{rot} = Density of air passing through rotameter, lbs/cuft.

ρ_F = Density of gas lbs/cu.ft.

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